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**COST OF PURIFYING  
MUNICIPAL WASTE WATERS  
BY DISTILLATION**

**AWTR-6**

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**U. S. DEPARTMENT OF HEALTH,  
EDUCATION, AND WELFARE**

**Public Health Service**

# **COST OF PURIFYING MUNICIPAL WASTE WATERS BY DISTILLATION**

by

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for

**The Advanced Waste Treatment Research Program**  
Basic and Applied Sciences Branch  
Robert A. Taft Sanitary Engineering Center

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**November 1963**

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Public Health Service Publication No. 999-WP-6

## ADVANCED WASTE TREATMENT RESEARCH PROGRAM

The Advanced Waste Treatment Research Program of the Public Health Service has two ultimate goals: One is to help abate our Nation's growing water pollution problems, and the other, more startling in concept, is to renovate waste water for direct and deliberate re-use.

Conventional water and waste treatment practices have little or no effect on many simple inorganic salts and permit an ever-increasing number and amount of highly complex, synthetic organic wastes to contaminate drinking water supplies. If we are to protect these water supplies for the future, it is mandatory that new technologies for water and waste treatment be developed.

Advanced Waste Treatment may be looked upon as a two-step process: (1) Separating concentrated contaminants from the purified water "product," and (2) disposing of the concentrated contaminants in a way that will render them forever innocuous. This "permanent disposal" of separated wastes is essential so that useable surface or ground waters will not be contaminated and so that the same contaminants need not be removed again and again from water supplies.

In one step, then, advanced waste treatment, i. e. , water renovation, could alleviate both water pollution and water supply problems — problems of increasing concern both in this country and in the rest of the world. The importance of advanced waste treatment was recognized by Congress in its 1961 amendments to the Federal Water Pollution Control Act (Public Law 87-88). This act directs the Secretary of Health, Education, and Welfare ". . . to develop and demonstrate practicable means of treating municipal sewage and other water-borne wastes to remove the maximum possible amounts of physical, chemical, and biological pollutants in order to restore and maintain the maximum amount of the Nation's water at a quality suitable for re-use."

It is too early to predict accurately the cost of advanced waste treatment technology, but the need is inevitable. Because of population increase and industrial expansion, we are approaching the time when our rivers and streams will no longer be able to assimilate our wastes and when there will be no more developable supplies of fresh water. Whatever the cost of this technology, one thing is certain — it must be the very lowest science and engineering can achieve. Anything more will impose an

inexcusable, perhaps intolerable burden on our country's economic growth and on its ability to compete in tomorrow's world.

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The author of this report, Dr. Jack A. Gerster, has for some time been interested in distillation and has published a number of papers in this field. In 1962 he was given the Professional Progress Award by the American Institute of Chemical Engineers for important contributions to research, especially in distillation, and to chemical engineering education.

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## ABSTRACT

Desk top studies of the cost of distillation of waste water are made with the types of equipment proposed for use with sea water. The estimates are based upon those for sea water, but are modified to conform with the differences in composition of sea water and waste water, and differences in the modes of operation for the two feeds. The smaller concentration of scaling materials in waste water should allow distillation to be carried out at a higher temperature than can be used with sea water. The smaller total contaminant concentration results in a smaller boiling-point elevation. Because of the need to dispose of the blowdown permanently it is necessary to concentrate to a far greater extent than in the case of sea-water distillation.

Three types of equipment: Multi-stage flash, multiple-effect, and recompression-flash, appear economically promising. For all types of equipment the cost for the distillation step alone is somewhat less than for sea water, but inclusion of costs for feed pretreatment and ultimate disposal of blowdown, bring the cost up to about that for sea water.

# **COST OF PURIFYING MUNICIPAL WASTE WATERS BY DISTILLATION**

## **SUMMARY**

An attempt has been made in this report to estimate the applicability of various distillation and evaporation processes for purifying municipal waste waters. The considerable experience gained through the Office of Saline Water in producing potable water from sea water by distillation has been utilized to choose desirable processes and to estimate costs for the purification of waste waters.

Several important differences exist between the two cases. In saline-water purification, the sea-water feed is roughly split equally into a distilled water product and a concentrated solution that is returned to the sea, while in waste-water purification, essentially all the water is removed as distillate from the solids that remain. Despite this high degree of water removal, the percent of dissolved solids in the blowdown will probably be less than for saline water.

This smaller concentration of solids results in a smaller boiling-point elevation than for sea water and in the possibility of greater steam economy. The variable nature of the solids in waste water affects the degrees of corrosion and scaling, which fix the maximum operating temperature. This temperature might be significantly higher for waste water.

A process is suggested for purification of municipal waste waters that includes a pretreatment to remove certain scale-forming inorganics, evaporation of 95 percent of the water by one of several economical vapor re-use procedures, and final evaporation of the remaining small amount of contaminant concentrate. The bulk of the report examines various vapor re-use procedures for the main evaporation step. Those appearing most desirable in terms of operability and cost are multiple-effect evaporation

combined with thermocompression evaporation, multi-stage flash evaporation, and recompression-flash evaporation. Economical sources of energy for the evaporation processes are examined, and the various problems are outlined that remain to be considered before any of the suggested processes can be adapted to the present application.

A preliminary cost analysis indicates that purification of waste by distillation, including pretreatment and disposal of the concentrated contaminant, requires about the same total expenditure as does separation of a sea-water feed into equal parts of distilled water and a 7 percent concentrate. The basis and assumptions used are summarized on pages 38 and 39.

## INTRODUCTION

### Background

The purpose of this study is to determine if the use of distillation as a means of removing small amounts of various organic and inorganic contaminants from municipal waste waters that are not removed by ordinary waste treatment is economically feasible.

The growing problem of these contaminants has been a major concern of the Division of Water Supply and Pollution Control of the U. S. Public Health Service, which has pointed out that, because various chemical products such as synthetic detergents and pesticides are comparatively resistant to modern waste treatment and to destruction by natural stream agencies, their concentrations in rivers, streams, and underground waters will increase as use of these substances increases. It is not known at what concentration in municipal water supplies these contaminants become a menace to public health, but a solution to the problem is desired before these concentrations are reached.

Spot tests made recently by the U. S. Public Health Service in various cities have shown that the effluent from an effective sewage treatment plant contains about 250 ppm more of inorganic solids than does the source of the city water supply and about 70 to 100 ppm more of organics (as COD).

The American Institute of Chemical Engineers has also shown interest in this area by its establishment of a Pollution Control Engineering Committee. This group has issued the first of a series of biennial surveys of recent developments in the field of water pollution control.<sup>7</sup>

### Nature of Contaminants

To cope with the water pollution problem, the Research Branch of the U. S. Public Health Service's Division of Water Supply and Pollution Control has undertaken to characterize the nature of contaminants and to study various methods for removing them. Much progress has been made in the characterization studies.<sup>5</sup> Although many contaminants have been identified (proteins, carbohydrates, tannins, lignins, and ether-extractable organics, for example), their number is very large and many remain unidentified.

The 1962 edition of the Public Health Service Drinking Water Standards reduces maximum limits for arsenic, copper, fluoride, lead, selenium, and zinc concentrations. New to the standards are limits on the concentrations of strontium-90,

radium-226, alkyl benzene sulfonate (to test detergent level), barium, cadmium, cyanide, nitrates, synthetic organic chemicals, and manganese, and a limit on the amount of beta activity. Since most of these substances are soluble and not ordinarily removed from drinking water, it is important to direct attention to procedures for their removal.

## Basis for Design of Distillation Process

Because the equipment requirements for separation by distillation are well known and available from the literature, this report can focus on an economic analysis. This analysis assumes that the amount of water to be treated is large — about 5 to 100 million gallons per day (mgd). It assumes also that the amount of contaminant is small, from 400 to 1,000 ppm, and that removal of essentially all the contaminants is desired. Any distillation process will, in fact, remove all nonvolatile contaminants in a single evaporation step if entrainment is kept low. Volatile contaminants may be removed in a preliminary evaporation step, but if their concentration is so small that their presence in the final product is not objectionable, this step with its added cost can be eliminated. It is furthermore assumed that the removed contaminants are recovered as a concentrate containing 2 percent solids, which means removal of 95 percent of the water. Most of the procedures considered are concerned with this step, because if low cost cannot be achieved here, distillation procedures may then be judged uneconomical. It is in this final step of the process that small amounts of concentrate are evaporated nearly to dryness for ultimate disposal.

Another important assumption is that temperature levels as high as 350° F can be tolerated in the evaporation equipment. Considerations for the choice of this value are given on page 17.

Finally, an important factor in the design of the process is the boiling-point rise of feed water. In the first step the effect of the contaminants upon the boiling point of the water is almost negligible, but as the feed water is concentrated by evaporation to 2 percent solids by some vapor re-use procedure, its boiling point becomes slightly elevated. According to Clark et al.,<sup>34</sup> water with a 2 percent sea-salts content has a boiling-point elevation of about 0.4° F. Only in the final step of the process, where the 2 percent contaminant concentrate is evaporated to dryness, will the boiling point change significantly, depending upon the nature of the contaminants.

## Distillation Processes Available

Many distillation processes are available that may be modified for use in the present application. Some are perhaps better known as evaporation processes. This report considers all processes whereby water is removed from the feed by evaporation or distillation followed by condensation.

### FACTORS AFFECTING PROCESS DESIGN

The general problem of purification of water by distillation is affected by three main factors: (1) Supply of necessary heat energy, (2) procedures for conservation or re-use of heat energy, and (3) design of heat exchange equipment. Each will be discussed briefly, then considered in detail.

#### Supply of Necessary Heat Energy

Most distillation or evaporation processes utilize steam as the source of heat energy, but since the production of steam can be costly, all other possibilities need to be considered also. For example, solar energy may be practical if equipment limitations can be removed; nuclear energy may be practical in large installations. Low-cost steam may result if the distillation purification plant can be made part of an electricity-generating plant, with exhaust steam from the turbines available for the distillation process. Another possibility is the elimination of all costs and inefficiencies of the steam generation step by evaporation of the contaminated water through use of submerged-combustion techniques. These techniques provide fuller use of available heat from the fuel, but introduce added equipment problems.

#### Procedures for Conservation or Re-use of Heat Energy

Unless provisions are made for conservation of the heat energy supplied, the cost of a distillation process becomes prohibitive. Heat can be conserved by use of the evaporated water as a source of heat in another chamber or vessel that operates at a slightly lower pressure, and thus at a slightly lower boiling temperature. The procedure may be repeated many times until the increase in equipment cost overtakes any added savings in steam cost. This principle has found wide use in multiple-effect evaporators.

An alternate procedure to operating the second chamber at a pressure lower than the first is to pass the water vapor into a compressor; this raises its temperature and pressure, and permits it to be used as the heating medium in the second chamber

without any lowering of the pressure there. Actually, there is no need in this instance to have a second chamber at all, since the compressed vapor from the first chamber may be used as the heat source for the first chamber. This design is known as a thermocompression or recompression evaporator. All the necessary input energy requirement for this process may then be mechanical rather than thermal (except during startup).

Flash evaporators also permit heat to be re-used and transferred under low temperature differences, a condition necessary for high thermodynamic efficiency. In this scheme hot feed water is allowed to flash or evaporate by flowing through a succession of chambers each at a slightly lower pressure than the previous one. The feed water is heated when it is allowed to act as the condensing medium for the flashed vapors obtained in each chamber; obviously some additional heat must be supplied from an external source, but the heat economy is good.

#### Design of Heat Exchange Equipment

Although the schemes just mentioned may permit as much as 10 to 30 pounds of water to be evaporated per pound of steam supplied, the equipment required is much more elaborate than if heat conservation were not employed. This fact requires careful and economical design, especially of the heat exchange equipment. For example, high heat transfer rates can be realized through use of long-tube, vertical, falling-film evaporators or through use of rotating surfaces to spread the liquid into a thin film. Optimum equipment design is considered in a later section of this report.

### **Information from Saline-Water Conversion**

Considerable information has been made available over the past several years on procedures for producing potable water from sea water or brackish water.

#### **OFFICE OF SALINE WATER**

The Office of Saline Water (OSW), a division of the U. S. Department of the Interior, has sponsored research programs and built demonstration plants to obtain exact cost information for a number of saline-water conversion processes.<sup>31</sup> Some of the most economically attractive of these proposed processes involve distillation.

## EXISTING SALINE-WATER PLANTS

Most of the fresh water now being produced from sea or brackish water is by distillation (Table 1).<sup>28</sup>

Table 1. PLANTS NOW PRODUCING FRESH WATER BY DISTILLATION OF SALINE WATER

| Location                                   | Capacity, mgd | Type              | Manufacturer                  |
|--------------------------------------------|---------------|-------------------|-------------------------------|
| Kuwait (Persian Gulf)                      | 2.5           | Multiple-effect   | G. & J. Weir and Westinghouse |
| "                                          | 2.5           | Multi-stage flash | Westinghouse                  |
| "                                          | 0.375         | "                 | Richardson                    |
| Aruba (Caribbean)                          | 3.5           | Multiple-effect   | G. & J. Weir                  |
| Curacao (Caribbean)                        | 1             | "                 | "                             |
| Nassau (Bahamas)                           | 1.2           | "                 | "                             |
|                                            |               | (under contract)  |                               |
| Kingley Air Base (Bermuda)                 | 0.225         | Recompression     | Cleaver-Brooks                |
| Dharan Air Base (Arabia)                   | 0.2           | "                 | "                             |
| Pacific Gas and Elec., (Morro Bay, Calif.) | 0.15          | Multiple-effect   | Lummus Co.                    |
| Isle of Guernsey                           | 0.6           | Multi-stage flash | G. & J. Weir                  |
| S. Cal. Edison (Oxnard, Cal.)              | 0.1           | "                 | Cleaver-Brooks                |
| Aircraft carriers (U. S., each)            | 0.2           | "                 | All manufacturers             |
| Shell Refinery (Cordon, Venezuela)         | 1.2           | "                 | —                             |

Additional information on the sea-water distillation plant on the Island of Aruba has been made available by Hubbell.<sup>17</sup> This plant employs five units to produce water containing 5 ppm of total solids from sea water containing 36,000 ppm of total solids. Each unit uses six-effect, submerged-coil distillation. The plant produces about 11,000 kilowatts of electric power. The boilers are oil-fired and produce steam at 850 pounds per square inch gage (815° F). The expected water cost is about \$1.75 per 1,000 gallons with some added credit for the power produced. The plant cost was \$11 million.

## OSW RESEARCH AND DEVELOPMENT

Small-scale research on various distillation processes by the Office of Saline Water includes projects at Polytechnic Institute of Brooklyn and Food Machinery and Chemical Corporation on vapor reheat, multi-stage distillation; and at General Electric and Bureau of Ships on wiped-film evaporators. A pilot plant using long-tube, vertical evaporators has shown the feasibility of scale control in saline-water evaporation and of the use of mild steel as a construction material. Solar distillation has also received attention. Possibilities for applying the results of these studies to the present problem are discussed later.

The Office of Saline Water was authorized in 1958 to build five demonstration plants for saline-water conversion, to obtain cost and performance data for the most promising processes. The five processes are as follows. <sup>31</sup>

1. Multiple-effect, long-tube, vertical (LTV) process. The sea-water plant at Freeport, Texas, produces 1 mgd of fresh water in 12 effects of the falling-film type. The maximum temperature is 250° F and the sea water is concentrated about four times.
2. Multi-stage flash distillation process. A second sea-water conversion plant, being built near San Diego, California, will also have a 1 mgd capacity.
3. Vapor recompression distillation process. At Roswell, N. Mex., is a plant with a design capacity of 0.25 to 1 mgd. It operates on brackish water containing 10,000 to 20,000 ppm of solids and reduces the solids content to less than 500 ppm.
4. An electrodialysis process using membranes.
5. A freezing process.

## Differences Between the Saline-Water Problem and the Present Problem

Differences between the saline-water problem and the problem of this study prevent the conclusions reached in the saline-water program from being applied directly to the problem of removing relatively small quantities of contaminants from water supply sources. These differences are:

1. In the saline-water program, the feed stream is usually separated into a concentrated salt solution and pure water. The concentrate is readily discarded to the sea and may consist of 50 percent or more of the feed. In the present problem there is no opportunity to discard large quantities of the feed; rather, the

feed must be separated into a fairly pure water and a very small quantity of concentrated contaminant.

2. The solids content of sea water, which is  $\sim 3.5$  percent, is greater than that of the feed in the present instance, which probably will not exceed 1,000 ppm (0.1%). The sea-water concentrate returned to the ocean in most of the saline-water processes contains about 7.0 percent solids, while in this report, the final contaminant concentrate from the vapor re-use equipment is assumed to contain 2 percent solids. These differences alter the degree of corrosion, boiling-point elevation, and fouling.

Corrosion is not greatly altered as the salt concentration in water is reduced, according to Fink.<sup>35</sup> Thus one might not expect appreciably lowered corrosion rates in waste-water evaporators. In the absence of dissolved oxygen, however, a condition approached within evaporators operating at elevated temperatures, steel is a possible material of construction in sea-water evaporators;<sup>30</sup> its use in the present application appears promising, but this promise should be validated experimentally.

The elevation in boiling point of water containing dissolved solids is proportional to the amount of these solids present. Thus in the present case the boiling-point elevation is less than that in saline water, and steam generated from waste waters is superheated to a lower degree than steam from saline-water evaporators, which permits greater thermodynamic efficiencies (p. 10).

The degree of fouling to be expected in contaminant evaporators is difficult to predict because of the wide variation in feed composition. Organic contaminants such as detergents probably will give little trouble in this regard, but the effects of other types of organic contaminants will have to be determined experimentally. Ground and river waters normally contain from 50 to 250 ppm of inorganic contaminants, consisting mainly of calcium and magnesium bicarbonates, and some sulfates. Fouling from deposited carbonates and sulfates could give trouble in contaminant evaporators, and some pretreatment of the evaporator feed may be necessary. Procedures commonly employed in boiler feedwater treatment might be considered: hot-process softening by lime or ion exchange. Both procedures would add to the cost of the operation. Monet has shown that the full cost of removal of 200 ppm of  $\text{CaCO}_3$  from 300 gallons of raw water per minute ranges from \$0.08 to \$0.19 per 1,000 gallons by ion exchange.<sup>36</sup> In the present application, full removal of all the dissolved solids in a pretreatment is not necessary; removal of inorganic solids

should be carried out only to a degree sufficient to prevent solids deposition. Obviously an optimization between the degree of pre-treatment and the solids content of the contaminant concentrate (assumed to be 2 percent, but subject to change) should be undertaken, but this is beyond the scope of the present report.

## Thermodynamics of Separation by Distillation

Application of thermodynamic principles to the present problem shows that if the boiling point of the waste water is only differentially higher than that for pure water, the minimum work requirement for the separation of water from contaminants under ideal, reversible conditions is zero. This condition cannot be realized, of course, because of the practical need for a finite rather than a differential value for temperature difference ( $\Delta t$ ) in the evaporation and condensation steps. Only with a finite  $\Delta t$  does the size of the equipment become practical. The thermodynamic analysis does show, however, that with very close approaches (small values of  $\Delta t$ ) in heat exchangers, with small pressure drops in lines and equipment, and with maximum efficiencies in pumps and compressors, the heat or work requirement for the distillation process may be quite small. The equipment size to achieve these idealized conditions may, however, be very large.

Dodge and Eshaya<sup>13</sup> have analyzed the thermodynamics of various desalting processes. For brackish water containing 5,000 ppm of NaCl at 25° C, the minimum work is 0.71 kilowatt-hour per 1,000 gallons of product for a 50 percent recovery of potable water, while the minimum work is 1.05 kilowatt-hours per 1,000 gallons for a 90 percent recovery. These figures are in contrast to a minimum work of zero for water containing no contaminants (nearly the present case).

These authors also analyzed the work requirements for a 50 percent recovery of potable water from a feed containing 3.5 percent NaCl. They showed that the minimum work is 5.8 kilowatt-hours per 1,000 gallons, while for a vapor recompression cycle with a mean  $\Delta t$  of only 5° F in the main heat exchanger, a zero  $\Delta t$  when heat is exchanged between the feed solution and the solutions leaving, and a 70 percent efficiency of the compressor, the work requirement is 38.5 kilowatt-hours per 1,000 gallons. These calculations show that the small but reasonable allowances for the necessary driving forces — irreversible effects — have multiplied the ideal work by a factor of 6.

Their conclusions also apply to the present case. The magnitude of nearly all the work or heat input requirement for any

distillation process depends upon the magnitude of inefficiencies or irreversible effects, i. e. , upon the magnitude of the temperature and pressure drops, the heat losses, and the prime-mover efficiencies, rather than upon the magnitude of the contaminant concentration. Because, however, the minimum work is 5.8 kilowatt-hours per 1,000 gallons in desalination while it is nearly zero in contaminant removal, the power requirement for waste-water distillation (in the first step of the process where 95 percent of the water is recovered) might be approximately  $(38.5 - 5.8)/38.5$ , or 85 percent of that for a sea-water feed.

## SOURCES OF ENERGY FOR DISTILLATION PROCESSES

As mentioned earlier, an important cost item in the distillation process is that associated with the energy requirement. Various sources of energy are now examined.

### Conventional Steam Plants

Steam from conventional boiler plants is widely used as an energy source for distillation processes. According to Nelson,<sup>24</sup> 1.77 million Btu (gross) are required to make 1,000 pounds of steam in an efficient boiler plant. If the steam pressure is within the normal range—between 125 and 600 pounds per square inch gage—the cost varies less than 10 percent with pressure. Steam cost is mainly then a function of the fuel cost (Table 2).<sup>25</sup>

Table 2. RELATION BETWEEN FUEL AND STEAM COST

| Oil cost<br>\$/bbl | Gas cost,<br>\$/1,000 std ft <sup>3</sup> | Steam cost,<br>\$/1,000 lb |
|--------------------|-------------------------------------------|----------------------------|
| 1.92               | 0.25                                      | 0.53                       |
| 2.24               | 0.29                                      | 0.62                       |
| 2.56               | 0.33                                      | 0.71                       |
| 2.88               | 0.37                                      | 0.79                       |

The price of fuels varies widely with location. Nelson<sup>23</sup> indicates that in 1956 the price of Bunker C fuel oil averaged \$2.14 per barrel on the Gulf coast or in California, \$2.76 in New York Harbor, and \$3.53 in central Michigan. If the price index of fuel oil in 1956 is taken as 100, the variation from year to year has been from 78 in 1950 to 123.8 in 1957, and prices have been generally upward since then. Gaseous fuels may cost as little as \$0.098 per 1,000 standard cubic feet at the well in southwest Texas to as much as three or even four times that value in the East.

In March 1956, the Office of Saline Water<sup>29</sup> suggested that steam costs be taken as 55 cents per 1,000 pounds. Steam at 160 pounds per square inch gage was supplied in 1961 by Dow to the Freeport, Texas, LTV demonstration plant at 45 cents per 1,000 pounds.<sup>30</sup> Higher costs are common in the East, however, where values of 80 cents to \$1 per 1,000 pounds are commonly employed in cost analyses.

### **Exhaust Steam From Turbines**

Brice and Townsend<sup>4</sup> state that "power generation is a necessary part of the economics of low-cost water production utilizing a fossil-fueled boiler." They are referring to the desirability of first passing high-pressure steam from a conventional plant through steam turbines to generate electricity, and then of using the turbine exhaust steam as the heat source in the distillation process.

Nelson<sup>26</sup> states that if the main boiler is operated at 1,000 pounds per square inch gage and the exhaust from the steam turbine driver for the electric generator is at 100 pounds per square inch gage, it is possible to generate 49 kilowatt-hours per 1,000 pounds of steam. If the value of the electric power is \$0.008 per kilowatt-hour, then the 49 kilowatt-hours has a value of 39 cents. If the steam cost is 80 cents per 1,000 pounds, then (80-39) or 41 cents is the cost of 1,000 pounds of 100 pounds per square inch gage steam for the distillation process. Brice and Townsend,<sup>4</sup> in their cost analysis of a large, multi-stage flash plant for sea-water conversion, use a steam cost of from 35 to 40 cents per 1,000 pounds when part of the steam cost is charged against power generation. Standiford and Bjork<sup>30</sup> report a steam cost of 40 to 45 cents per 1,000 pounds in an analysis of a multiple-effect evaporation plant.

### **Steam Generation in Nuclear Reactors**

Two studies have shown that the economics of steam generation from nuclear energy are favorable where a large amount of relatively low-temperature thermal energy is required in a single-purpose plant.<sup>2,3</sup> The smallest economical size of a nuclear steam generator is one that would provide energy both for evaporation and for driving a majority of the pumps in a plant producing 50 mgd of distilled water by the multi-stage flash evaporation process.<sup>6</sup> A 370-thermal-megawatt steam generator (net to the evaporator) is required in this application, and a pressurized light-water reactor appears to be the optimum type. With this design, steam is estimated to cost 37 cents per 1,000 pounds including all costs associated with the nuclear steam

generator. This cost figure is no more favorable than that given above for ordinary steam when part of the steam cost is charged against power generation. It is likely, however, that 10 years hence the cost of ordinary fuels will be increased while the cost of nuclear steam generation, owing mainly to technological advances, will be lower.<sup>4</sup> Another factor favoring the nuclear steam source is that no power generation is involved; efficient tying together of power demands and demands for distillation may not always be possible.

### Submerged Combustion as a Heat Source

In the submerged-combustion technique, fuel and air are supplied to a burner located under the surface of the liquid to be evaporated; the flame and hot combustion products come in direct contact with the water, and the percentage of available heat received by the process liquid is very large. According to Cronan<sup>11</sup> a typical unit can liberate 4 to 5 million Btu per cubic foot per atmosphere per hour. Limitations of the procedure include considerable entrainment, foaming, vibration, and surging of the liquid being evaporated. It is also necessary to employ gas as a fuel since no satisfactory burner has been developed that uses oil or powdered coal. Submerged combustion has been applied when the boiling liquid has been under pressures as high as 100 pounds per square inch absolute.

Experimental evaluation of a submerged-combustion unit for producing hot water was carried out by Kobe and Hall.<sup>19</sup> About 90 percent of the gross heating value of the natural-gas fuel was utilized in heating large quantities of water to 190° F in an enclosed 500-gallon tank. Application of submerged combustion in industrial waste treatment has been discussed by Weisman.<sup>32</sup>

Cronan<sup>11</sup> describes the installation at Jefferson Lake Sulfur Co., where 0.5 mgd of water is heated under pressure to 325° F by submerged-combustion techniques. Natural-gas burners release 45 million Btu per hour. The combustion products exhaust to a 10-million-Btu-per-hour reheater where additional gas flames boost their temperature to 800 to 1200° F. These gases then drive a gas turbine that supplies air to the burners; stack gases from the reheater and from the turbine exhaust are used to preheat the incoming feed water. Raw water is used, and no corrosion problems have been encountered.

The largest multiple-submerged-combustion installation uses 383 million Btu per hour to evaporate magnesium chloride solutions for Dow Chemical Co. at Freeport, Texas.

Use of submerged-combustion techniques as a heat source for multiple-effect or multiple-flash evaporators is feasible, especially for the latter where the water is heated without phase change in one chamber, then flashed in later chambers. Submerged combustion works best for heating water under pressure without evaporating it in the combustion chamber. If evaporation occurs, the water vapor is contaminated with the combustion products, making subsequent condensation difficult owing to the inert gases present.

The heat efficiency of submerged-combustion units is apparently not so much greater than conventional boilers that its disadvantages of high entrainment and surging are overcome. Its industrial applications have been mainly in situations where fouling and corrosion are controlling factors. In situations where part of the cost of conventional steam generation can be charged against electric-power generation, the cost of submerged combustion ceases to be competitive.

## Solar Distillation

The use of heat from the sun as an energy source in saline-water conversion has received much attention.<sup>21</sup> In most processes under study, very large basins of sea water are used to collect the heat, and the water is vaporized directly throughout the basin. Clear plastic-film barriers prevent vapor loss to the air but permit the solar radiation to be received. These barriers also act as the condensing surface. Some consideration has been given to operation at higher temperatures so that multiple-effect principles can be employed. Elimination of the plastic barrier and use of a nonvolatile organic liquid to cover the surface of the water in the basin to prevent loss of vapor have also been proposed; the heated water would be sent to a multiple-effect evaporator.

It might be supposed that, because steam costs constitute about 50 percent of the total costs in conventional distillation processes, solar distillation processes should receive serious attention. There are, however, a number of reasons why these processes do not appear to be economically feasible. In the conventional solar distillation process, the life of the plastic barrier is unknown, yet this is a major cost consideration. Operation at higher temperatures so that multiple-effect principles can be employed does not appear desirable, because radiation losses would be tremendously large. Also, in all solar processes considerable safety measures must be employed in the design owing to the unpredictable extent of solar radiation available from day to day.

Lof<sup>22</sup> has stated that the potential cost of producing pure water from sea water is \$1 per 1,000 gallons, based upon a maximum hourly receipt of solar energy of 300 Btu per square foot. Experience gained at the Florida Solar Distillation Research Station<sup>1</sup> has shown that a productivity attainment of 0.10 gallon per square foot per day at solar radiation intensities of 2,000 Btu per square foot per day is possible. Since 2,000 Btu can evaporate about 2 pounds or 0.24 gallon of water, only 42 percent of the available radiation is effectively utilized. For a plant to produce 10 mgd of purified water, a surface area of 100 million square feet is required; this is a basin 1.9 miles square (3.6 square miles). It is apparent that solar distillation procedures are both unwieldy and expensive, and so are not recommended for the present application.

## **SPECIFIC DISTILLATION PROCESSES**

On page 5, it can be seen that the three cost factors affecting water purification by distillation are: (1) Energy cost, (2) cost of effective re-use of steam, and (3) equipment cost. Energy cost has just been considered, and the latter two will now be considered together. It is desirable to treat together the cost of methods for re-use of steam and equipment cost because choice of the procedure for re-using the steam fixes the type of equipment, which in turn fixes the cost.

### **Cost of Saline-Water Conversion in Long-Tube, Vertical (LTV), Multiple-Effect Evaporators**

Calculation studies made for the Office of Saline Water using long-tube, vertical (LTV), multiple-effect evaporators are useful in considering the same type of equipment for purification of water containing 1,000 ppm of contaminants. One study<sup>30</sup> suggested the use of exhaust steam from a power plant and estimated the total cost of distilled water as 23 cents per 1,000 gallons. The capacity of the plant was taken to be about as large as could be provided as a single unit, i. e., 17 mgd of distilled water. Reduced efficiencies and higher labor costs for smaller plants made their water costs higher.

In another paper study the powerhouse turbine drove a vapor compressor, which permitted several LTV evaporator bodies to act as thermocompression evaporators, while the turbine exhaust was used to heat a separate multiple-effect evaporator having 10 bodies. No auxiliary power was generated in this scheme, so that all costs were chargeable to the production of water. Annual capital and operating costs (estimated in 1955) are shown in Tables 3 and 4 for a 16.65 mgd plant.<sup>30</sup>

**Table 3. ESTIMATED ANNUAL CAPITAL COSTS OF A 16.65 mgd, LTV EVAPORATOR SYSTEM FOR SALINE-WATER CONVERSION**

|                                              |             |
|----------------------------------------------|-------------|
| Boiler . . . . .                             | \$1,020,000 |
| Turbine-generator . . . . .                  | 170,000     |
| Turbine-vapor compressor . . . . .           | 925,000     |
| Evaporators . . . . .                        | 2,425,000   |
| Heat exchangers . . . . .                    | 991,000     |
| Pumps . . . . .                              | 138,000     |
| Piping . . . . .                             | 305,000     |
| Instruments . . . . .                        | 40,000      |
| <hr/>                                        |             |
| Total, installed process equipment . . . . . | \$6,014,000 |
| <hr/>                                        |             |
| Site development . . . . .                   | 70,000      |
| Office, shop, lab . . . . .                  | 135,000     |
| Weatherproofing . . . . .                    | 300,000     |
| Engineering . . . . .                        | 481,000     |
| Contingencies . . . . .                      | 600,000     |
| <hr/>                                        |             |
| Total plant cost . . . . .                   | \$7,600,000 |

**Table 4. ESTIMATED ANNUAL OPERATING COSTS OF THE LTV EVAPORATOR SYSTEM**

|                                             |             |
|---------------------------------------------|-------------|
| Interest, 3% of total capital . . . . .     | \$ 228,000  |
| Insurance, 1% of total capital . . . . .    | 76,000      |
| Depreciation, 5% of total capital . . . . . | 380,000     |
| Maintenance, 3% of equipment cost . . . . . | 180,000     |
| Labor, \$2.50/man-hr . . . . .              | 138,000     |
| Fuel, \$0.30 per million Btu . . . . .      | 991,000     |
| <hr/>                                       |             |
| Total . . . . .                             | \$1,993,000 |

Water cost is then \$0.328 per 1,000 gallons, based on 365-days-per-year operation.

#### **RECENT EXPERIMENTS WITH SALINE-WATER CONVERSION BY LTV EVAPORATORS**

Because the costs (Tables 3 and 4) were attractive, the Office of Saline Water authorized operation of a pilot plant at Wrightsville Beach, N. C., to confirm the operability of LTV evaporators. The pilot plant evaporator contained seven 2-inch tubes, each 24 feet long. Under conditions where the liquid is fed to the tops of the tubes and falls in a thin film throughout the

length of the tube while boiling takes place, the expected heat transfer coefficients were obtained; they varied from 350 to 400 at about 100° F to 700 to 800 at 250° F.<sup>30</sup> It was also found that scale could be prevented at temperatures below 250° F and that steel could be used as the construction material. These pilot plant results prompted authorization of a 1 mgd demonstration plant at Freeport, Texas. This plant uses a 12-effect, falling-film, LTV evaporator. The feed is de-aerated and preheated successively in two sets of heat exchangers through use of the heat from the steam condensate of each effect and from vapor bled from each effect. Forward feed is used, maximum and minimum temperatures are 250 and 120° F, and the sea-water feed is concentrated fourfold. The total water cost in the demonstration plant should be about \$1.04 per 1,000 gallons. All factors known to date indicate that a plant of reasonable size (over 15 mgd) could produce water for \$0.35 per 1,000 gallons by this type of operation.

### **Cost of LTV Evaporators for Purification of Waste Water**

Application of LTV evaporators to the present problem of waste-water purification would lead to distilled-water costs that are lower than those given on page 16 for a sea-water feed if the maximum temperature level could be raised above 220 to 250° F. This maximum temperature level in sea-water evaporators cannot be readily exceeded because of fouling and scaling difficulties. If it is assumed that the fouling problem in the present instance can be solved by the pretreatment procedure suggested on page 9, the main limitations on use of higher temperatures appears to be the added cost of the high pressures associated with these temperatures, and the possibilities of increased corrosion. Little information exists on corrosion rates of steel at elevated temperatures in water containing varied contaminants, but without appreciable dissolved oxygen. With at least a partial removal of inorganic solids for scale prevention, it is hoped that a 350° F temperature level could be tolerated. At this level the corresponding pressure is below 150 pounds per square inch gage, so that standard valves, piping, and equipment can be used throughout. In the remaining cost analyses of this report, a 350° F maximum temperature level is assumed; if higher temperatures could be tolerated, an additional cost reduction would result.

Consider a multiple-effect system in which the maximum and minimum temperatures are 350 and 120° F instead of the 250 and 120° F values upon which Tables 3 and 4 are based. This would permit a  $(350-250)/(250-120)$  or 77 percent increase in the number of effects without any increase in the steam requirement. In

Table 3, this would increase some of the capital costs. In the plant described on page 15, about 45 percent of the evaporation occurs in the 10-effect evaporator while 55 percent occurs in the thermocompression evaporator. Thus in a modification of Table 3 for the new situation, the costs of boiler, turbine generator, and turbine vapor compressor are unchanged, while the cost of the evaporators is increased by 77 percent, and the cost of other capital items is increased by 45 percent of 77 percent or 35 percent. These factors are approximate, but are helpful in arriving at an approximate water cost (Table 5).

Table 5. ESTIMATED ANNUAL CAPITAL COSTS OF LTV EVAPORATOR SYSTEM

|                                                            |              |
|------------------------------------------------------------|--------------|
| Boiler, turbine generator,<br>and turbine compressor . . . | \$ 2,115,000 |
| Evaporators . . . . .                                      | 4,300,000    |
| Other process equipment . . . . .                          | 1,990,000    |
| Total, installed process<br>equipment . . . . .            | \$ 8,405,000 |
| Other capital costs . . . . .                              | 2,140,000    |
| Total plant cost . . . . .                                 | \$10,545,000 |

The annual operating cost for an LTV evaporator system may now be estimated by use of the same percentage figures as in Table 4 for interest, insurance, depreciation, and maintenance; by assumption of a labor cost 35 percent above the previous value; and by assumption of a fuel cost 5 percent greater than the previous value, owing to lowered values for latent heat of evaporation and higher heat losses at the elevated temperatures employed (Table 6).

Table 6. ESTIMATED ANNUAL OPERATING COSTS OF LTV EVAPORATOR SYSTEM

|                                                                         |              |
|-------------------------------------------------------------------------|--------------|
| Interest, insurance, and depreciation, 9%<br>of total capital . . . . . | \$ 950,000   |
| Maintenance, 3% of equipment cost . . . . .                             | 252,000      |
| Labor . . . . .                                                         | 187,000      |
| Fuel . . . . .                                                          | 1,041,000    |
| Total . . . . .                                                         | \$ 2,430,000 |

The total in Table 6 is the annual cost for producing 77 percent more distilled water from the multiple-effect evaporators than before, and the multiple effects produced 45 percent of the total evaporation; thus in the new case the total water produced is  $(0.55) + (0.45)(1.77) = 1.347$  times greater than before. If the old water cost was \$0.328 per 1,000 gallons with an annual operating cost of \$1,993,000, then the new water cost is  $(\$0.328)(2,430,000/1,993,000)(1/1.347) = \$0.297$  per 1,000 gallons. This cost should not be construed as an absolute cost figure at today's prices, but merely indicates that waste water, suitably pretreated, can be separated into a distillate consisting of 95 percent of the feed and a contaminant concentrate at a cost that is approximately  $0.297/0.328$  or 91 percent of that for desalinating sea water.

#### SUPPLEMENTARY INFORMATION ON MULTIPLE-EFFECT EVAPORATOR DESIGN

Florine and Dedert, of the Swenson Evaporator Co., describe evaporator costs and list procedures for choosing the economical number of effects.<sup>15</sup> Costs of labor, water, power, and maintenance do not vary with the number of effects, so the calculation in Table 6 is slightly conservative. Kohlins and Englander<sup>20</sup> discuss the problem of whether to follow the ASME unfired-pressure-vessel code in the manufacture of evaporators, consider the relative merits of shop- vs. field-erected units, and describe construction techniques used for relieving stress.

Wiper blades have been used to improve heat transfer rates in vertical-tube evaporators.<sup>8</sup> The apparatus tested by the General Electric Co. for this purpose consisted of a 6-inch-diameter copper tube, 50 inches long, with a wall thickness of 0.050 inch, inside of which a wiper assembly rotates at 200 rpm about the vertical axis of the cylinder. Sea-water feed is distributed at the top of the inside of the tube and is spread by the wiper into a thin film; steam condensing on the outer surface of the tube causes the sea water to evaporate without boiling or bubbling. Many conditions of wiper design, speed, and load have been investigated, and tests are continuing under a grant from the Office of Saline Water.

The principle of operation here is similar to that employed in a "turbulent-film evaporator," in which the rotor shaft has blades that rotate within a minute clearance of the heated wall.<sup>33</sup> The main advantages of this type of evaporator are that contact times are low, there is little possibility of overheating, and there is no elevation in boiling point from hydrostatic head. These

advantages are useful in handling a heat-sensitive organic, but not in the present application. Inspection of the high cost figures for this type of evaporator<sup>33</sup> eliminates it from further consideration.

The heat transfer advantages obtained with forced-circulation evaporation are demonstrated by the studies of Eibling and Hyatt.<sup>14</sup> At a water velocity of 10 feet per second, the heat transfer rate in forced convection is about 2.5 times that obtained with natural convection; percentagewise, the effect of forced circulation increases as the  $\Delta t$  decreases. Dropwise condensation with tubes coated with Teflon plastic was also found to increase the heat transfer rate 13 to 33 percent over film condensation on brass, and the improvement is even greater when the condensing film  $\Delta t$  is very small, around 1 to 2° F.

## Multi-stage Flash Evaporation for Purification of Waste Water

Development of the multi-stage flash evaporation process for conversion of sea water to potable water has been extensive,<sup>4</sup> and is attractive because of the tremendous re-use of steam that is possible. In one design, 13.9 pounds of fresh water are produced per pound of steam. The process should have excellent economic feasibility for treatment of waste waters.

### PROCESS DESCRIPTION

A typical flow diagram for purification of contaminated water is shown in Figure 1, and a detail of a single evaporator is shown in Figure 2. With reference to the latter figure, the

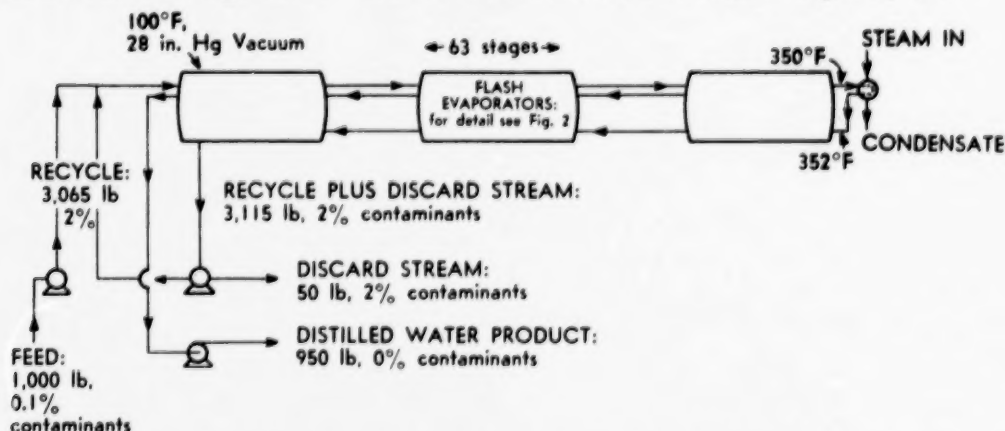


Figure 1. Flow diagram for purification of waste water by multi-stage flash evaporation.

principle of operation may be described. The contaminated feed flows through the inside of the condenser tubes, where it is pre-heated by taking up the heat of condensation of the evaporated water condensing on the outside of the tubes. After the feed is

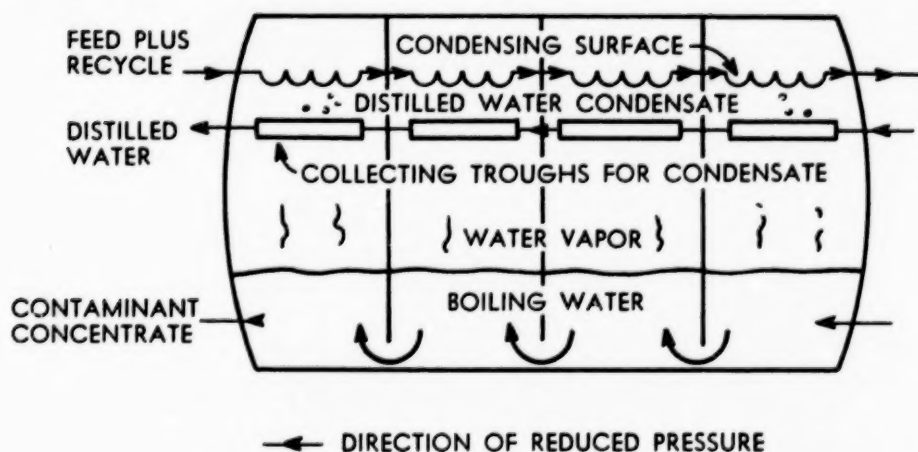


Figure 2. Detail of flash evaporator containing four effects or stages.

preheated to  $350^{\circ}\text{F}$  by successive passage through each of the stages, it is heated an additional few degrees by steam ( $2^{\circ}\text{F}$  in the design shown in Figure 1) then returned to the shell side of the highest temperature stage. Up to this point the feed has remained entirely liquid, which requires its pressure to be above 120 pounds per square inch gage.

The shell side liquid now cascades from stage to stage, each of which is at a slightly lower pressure than the previous one; as the liquid is reduced in pressure, a small amount flashes. The flashed vapor condenses on the outside of the feed preheater pipes, below which the condensate is collected in troughs and allowed to flow successively along in the same direction as the shell side liquid. At the lowest pressure stage the condensate is pumped out as purified product, while the unevaporated liquid is partly recycled and partly removed as a contaminant concentrate to be evaporated to dryness in a batch evaporator. The lowest pressure stage is held at as low a pressure as practical, probably a vacuum of 28 inches of Hg, corresponding to a boiling temperature of  $100^{\circ}\text{F}$ . A de-aerator is employed for the feed at the lowest pressure stage, where inerts are removed by an ejector system.

Figure 1 shows the relative stream quantities and concentrations for each 1,000 pounds of feed. The feed contains 999 pounds of water and 1 pound of contaminants (1,000 ppm), which is separated into a distilled water product containing 950 pounds of pure water and a contaminant concentrate containing 49 pounds of water and 1 pound of contaminants (2 weight percent contaminants). With a total temperature drop of the shell side fluid of  $352-100$ , or  $252^{\circ}\text{F}$ , and a temperature drop in each stage of  $4^{\circ}\text{F}$ , a total of 63 stages are required. If  $T$  is the average pounds of throughput of the shell side fluid per 1,000 pounds of

fresh feed, and the average value for the latent heat of evaporation is 952 Btu per pound, the pounds of liquid evaporated per stage averages  $4T/952$ , and the total evaporation is  $(4T/952)(63)$ . Equating this value to the desired evaporation, 950 pounds, shows that  $T$  must be, on the average, 3,590 pounds, and the evaporation per stage is 15.1 pounds. The initial throughput must be 4,065 pounds, made up of 1,000 pounds of fresh feed and 3,065 pounds of contaminant concentrate; the liquid discharge from the low-pressure stage consists of 3,115 pounds and contains 2 weight percent contaminants; the liquid to the first stage contains 1.54 weight percent contaminants.

The optimum mechanical design and operating conditions have been carefully considered for a sea-water feed.<sup>4</sup> Optimum mechanical features include condenser tubes that run continuously through several stages, perhaps as many as seven, with tube sheets employed only at the ends. Baffles are used instead of tube sheets for the other stage-to-stage separations, leakage being minimized either by close fit or by sealing. The evaporator vessels for a 25 mgd plant would be 30 feet in diameter; they are horizontal cylinders designed to take full advantage of their shape to save metal. Steam-driven turbines are specified as drivers because with the large quantities of low-pressure steam required in the process (the maximum brine temperature is around 220° F because of corrosion difficulties) these drivers save several cents per 1,000 gallons of product over electric motors.

Optimum process design includes choice of number of stages and choice of the terminal temperature difference (TTD). Unlike a multiple-effect evaporator, these two factors may be chosen independently.

#### CHOICE OF NUMBER OF STAGES

The number of stages is fixed by two factors: the pressure differential between stages, and the maximum and minimum temperatures that can be accommodated at the two ends of the stages. Because a given stage-to-stage pressure differential is required for satisfactory regulation of shell side brine flow, a minimum temperature difference is required from stage to stage, depending upon the absolute pressure. This minimum difference is apparently about 2° F.<sup>4</sup> The temperatures at the ends of the stages are limited by maximum practical vacuum on one end and maximum practical pressure at the other. For purification of contaminated water, these values are 28 inches of mercury vacuum and 150 pounds per square inch absolute since both of these conditions are

readily obtainable without use of special equipment. These pressures correspond to temperatures of 100 and 358° F, respectively. Thus in the present application where a 258° F total drop and a 2° F drop per stage are possible, as many as 129 stages may be used; the economical number of stages, as discussed later, may be less than 129.

Study of any individual stage shows that if the shell side liquid leaving a stage is 2° F lower than the entering shell side liquid, then the fluid inside the tubes is heated exactly by 2° F. The only assumptions involved in this statement are equality of liquid-phase heat capacities for the shell and tube side liquids and lack of heat loss; both are valid assumptions. The proof results from the fact that the sum of the shell side liquid and condensate flow rates equals the tube side flow rate, so that if the first two streams lose 2 Btu per pound of total flow, the latter stream must gain 2 Btu per pound.

An insight is gained into the value of having as small a temperature drop between stages as possible for the shell side liquid by a comparison of the two cases shown in Figure 3. In both

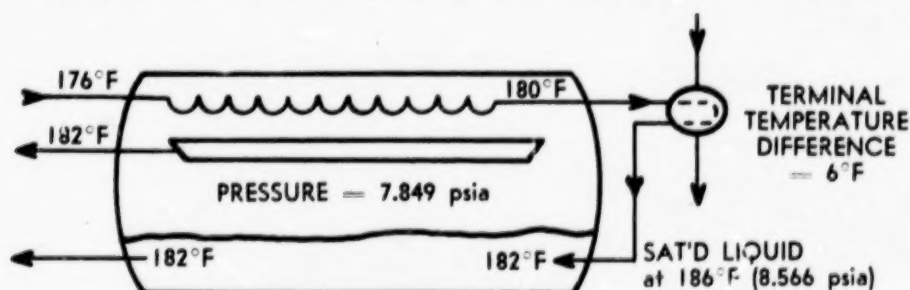
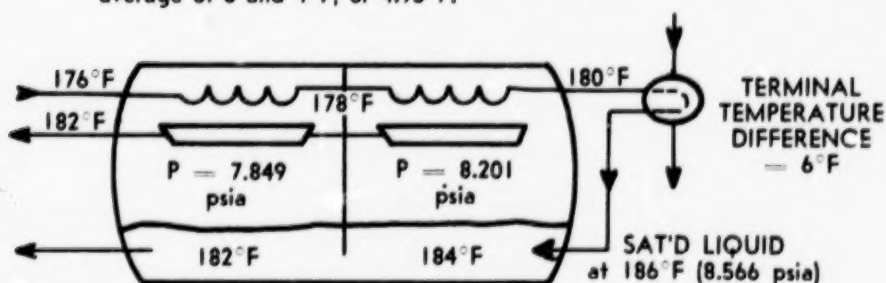


Figure 3. Case 1. (Top) Flash evaporator for transferring 4 Btu per pound of entering saturated liquid. Log mean temp difference is log mean average of 182-180 and 182-176, or 3.64°F.

Case 2. (Bottom) Flash evaporator (2 stages) for transferring 4 Btu per pound of entering saturated liquid. Log mean temp difference is log mean average of 6 and 4°F, or 4.95°F.



the TTD is 6° F ( i. e. , the tube side liquid is raised 6° F by steam before being returned to the stages as shell side liquid), but in the first case the required heat transfer area is 4.95/3.64,

or 1.36 times greater than in the second. This advantage is offset somewhat by the need to provide an additional stage in the second case. Thus for any given value of TTD, an economic balance is required to determine the optimum temperature drop between stages, which then fixes, for a given set of end conditions, the optimum number of stages.

#### CHOICE OF TERMINAL TEMPERATURE DIFFERENCE (TTD)

The previous paragraph showed that, for any given value for TTD, an economic optimum number of stages exists for a given set of end conditions. An alternate approach to the problem is to calculate, for any given number of stages, the optimum value for TTD. In this instance very small values for TTD produce very low steam costs, but, because of the low temperature differences, also produce very large costs for condenser tubes within the stages. On the other hand, very large values for TTD produce high steam costs and low equipment costs; thus for any given number of stages, an economic optimum exists for TTD.

#### OPTIMUM NUMBER OF STAGES AND TTD FOR SEA-WATER EVAPORATION

Economic balance calculations were done for sea-water evaporation by Brice and Townsend<sup>4</sup> to determine the optimum number of stages and optimum TTD. For a given number of stages, say 60, a minimum water cost was found as a function of TTD. The procedure was repeated for other total numbers of stages, and the minimum water cost for all cases considered was obtained when the number of stages was 50 and the TTD was 4° F. Since the terminal temperatures were 220 and 90° F, the temperature drop per stage of the shell side fluid was 2.6° F. The same result was obtained for outlet salt concentrations, which varied from 1.5 to 2.0 times that for sea water.

The tube brine velocity, 5 feet per second, was selected as a value sufficiently high to give good heat transfer coefficients but low enough to keep pumping costs and corrosion-erosion values reasonably low. Another study by the same authors<sup>4</sup> indicated that maximum temperatures higher than 220° F were economically desirable if corrosion could be minimized, but that minimum temperatures below 90° F were not desired because of the rapidly increasing volumes of vapor to be handled as the temperature (and especially the pressure) is reduced below this value. A summary of the costs of producing pure water from sea water by the plant just described is given below.

Table 7. WATER COST FOR OPTIMIZED PLANT

|                                                            |        |                 |
|------------------------------------------------------------|--------|-----------------|
| Capacity, mgd . . . . .                                    |        | 49.3            |
| Stages . . . . .                                           |        | 52              |
| TTD, °F . . . . .                                          |        | 4               |
| Lb water produced/lb steam. . . . .                        |        | 13.65           |
| <hr/>                                                      |        |                 |
| Total capital cost of plant, thousands of dollars. . . . . |        | 30,700          |
| <hr/>                                                      |        |                 |
| Operating costs                                            | \$/day | Cents/1,000 gal |
| <hr/>                                                      |        |                 |
| Electric power . . . . .                                   | 912    | 1.85            |
| Steam (via nuclear steam generator) . . . . .              | 11,090 | 22.49           |
| Chemicals. . . . .                                         | 1,723  | 3.50            |
| Supplies and maintenance materials . . . . .               | 246    | 0.50            |
| Operating labor . . . . .                                  | 314    | 0.64            |
| Maintenance labor . . . . .                                | 185    | 0.37            |
| Payroll extras . . . . .                                   | 83     | 0.17            |
| Overhead . . . . .                                         | 57     | 0.12            |
| Amortization at 4% . . . . .                               | 5,915  | 12.00           |
| Interest on working capital. . . . .                       | 144    | 0.29            |
| Real estate (\$5,000/acre at 4%) . . . . .                 | 48     | 0.10            |
| <hr/>                                                      |        |                 |
| Less tube salvage value . . . . .                          | (129)  | (0.26)          |
| <hr/>                                                      |        |                 |
| Total operating cost . . . . .                             | 20,588 | 41.77           |

Table 7 shows a water cost of 42 cents per 1,000 gallons with steam produced from a nuclear steam generator. With conventional steam, the water cost is lowered to 35 cents per 1,000 gallons with part of the steam cost defrayed from power generation, on the assumption that the power plant would be base loaded the same as the water plant. It is also stated that improvements in nuclear steam generation should permit this plant to produce water in 1972 at a cost of from 24 to 31 cents per 1,000 gallons.

#### COST OF PURIFYING WASTE WATER

The cost of producing distilled water from a 1,000 ppm waste-water feed in a similar multi-stage flash plant would be lower than for saline-water conversion if the plant is operated with a higher maximum temperature and thus with a larger number of stages. It is beyond the scope of the present report to compute the cost of an optimized plant for treating contaminated water, since great effort involving an electronic computer solution was required to optimize the saline-water design. If, however, the same optimum values of 4° F TTD and a 2.6° F

temperature drop per stage found for the saline-water plant are used for the waste-water purification plant, an approximate cost can be computed.

With a total temperature difference of  $358-100$ , or  $258^{\circ}\text{F}$ , as discussed on page 24, the number of stages is  $258/2.6$  or 99. Consider now a plant having the same circulation rate as that in Table 7. The electric power cost (for circulation) and steam cost will be about the same as in Table 7, while the cost for chemicals will be reduced, probably not exceeding 0.50 cents per 1,000 gallons. All other costs in Table 7 except electric power, steam, and chemicals will be increased owing to the increased number of stages; the cost per stage will not vary appreciably since the Btu transferred per stage will be about the same. In Table 7 this cost per stage is 0.268 cents per 1,000 gallons; for 99 stages, the cost is 26.5 cents per 1,000 gallons.

Table 8. SUMMARY OF COSTS FOR WASTE-WATER PURIFICATION PLANT

|                           |             |
|---------------------------|-------------|
| Electric power . . . . .  | 1.85        |
| Steam . . . . .           | 22.49       |
| Chemicals . . . . .       | 0.50        |
| All other costs . . . . . | 26.50       |
| <hr/>                     |             |
| Total . . . . .           | 51.34 cents |

The costs in Table 8 are for the production of purified water, which, because of the increased number of effects, is greater in amount by  $99/52$ , or 1.90. Thus the costs on the basis of 1,000 gallons are, in total,  $51.34/1.90$  or 27 cents. As mentioned previously, this figure should not be considered an absolute value, but rather an indication that, with an increase in the number of stages, purification of contaminated water can be accomplished at a cost that is  $27/42$  or 65 percent of that for desalination of sea water, provided, of course, that the assumptions made above are valid.

#### PROCESS VARIATION

A slightly different multi-stage flash evaporation process has been studied by Othmer et al.<sup>27</sup> In this process the flashed vapor, rather than transfer its enthalpy when condensed to the feed, transfers it to the recirculated distilled water by direct contact. The heated distillate then transfers its heat in an exchanger to the feed. This procedure would seem to lead to higher equipment and operating costs than the multi-stage flash process described above because of the engineering difficulty of condensing

vapors directly into a liquid stream, and because of the need for external heat exchangers.

## Recompression Evaporation for Purification of Waste Water

### PROCESS DESCRIPTION

Another method of saving heat through vapor re-use is employment of a recompression or thermocompression evaporator. In this apparatus the evaporated water vapor is raised slightly in pressure and enthalpy and used as the heating medium for the evaporation. Under conditions of small temperature difference between the boiling water and the heating medium, the quantity of heat added can be kept small. This scheme is not attractive with high boiling-point elevations, but this limitation does not apply here.

The vapor from the evaporator may be raised in pressure and enthalpy by means of a mechanical compressor, or by a steam jet recompressor (see Figures 4 and 5). With the former

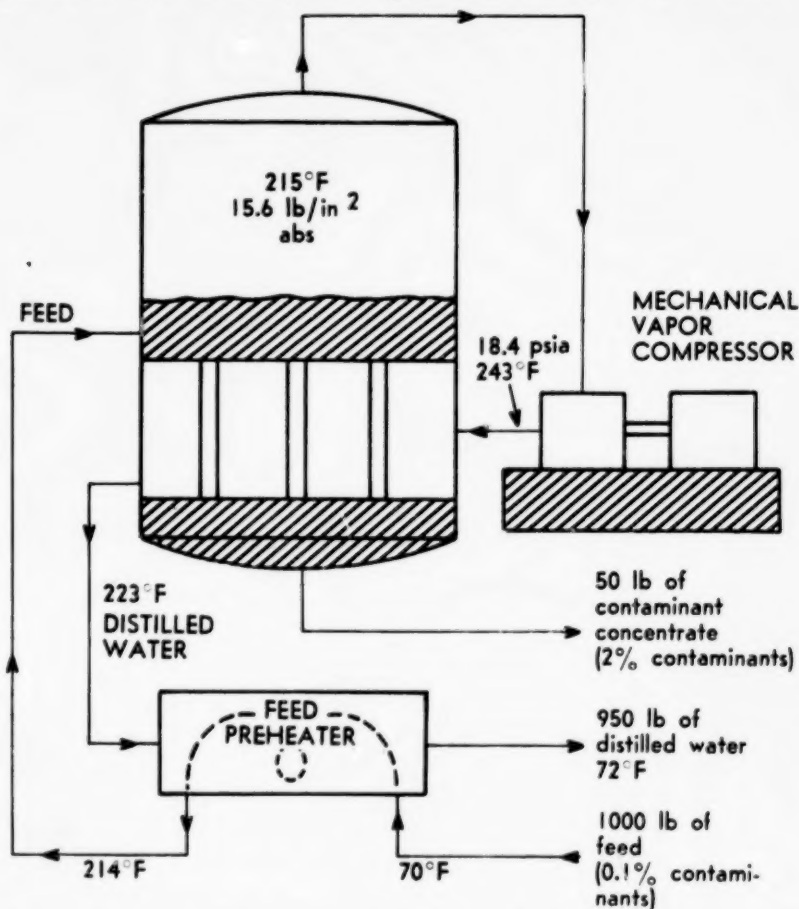


Figure 4. Recompression evaporator using mechanical vapor compressor.

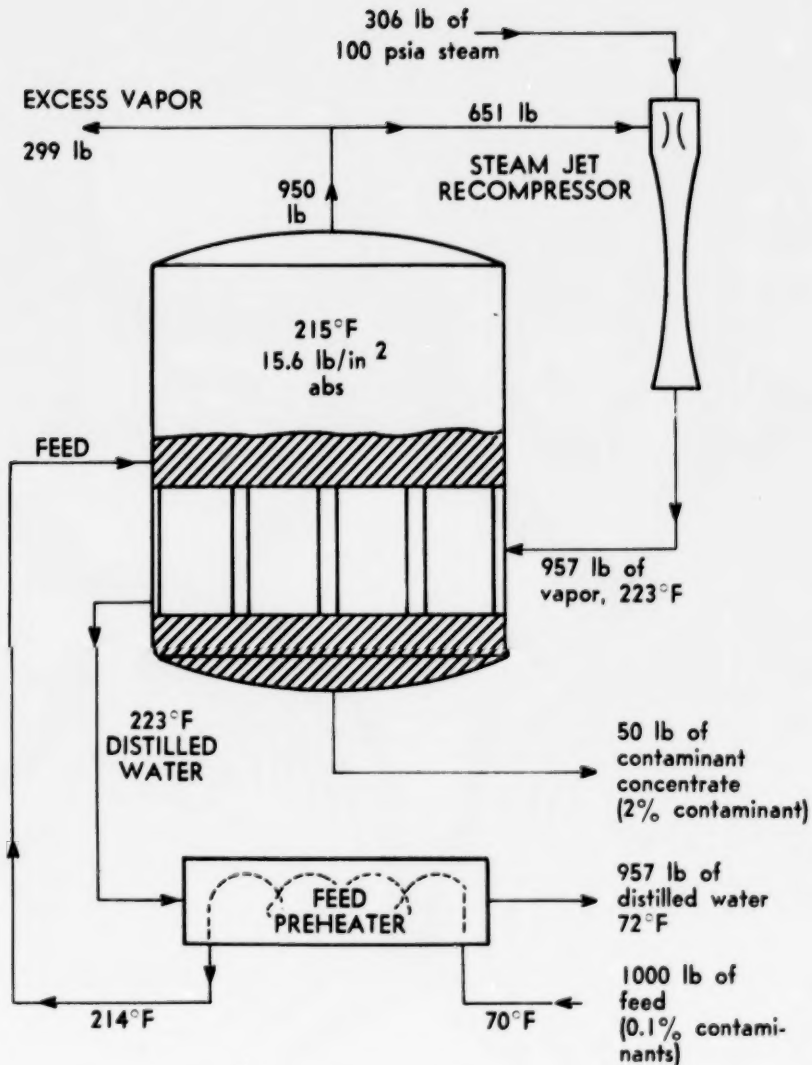


Figure 5. Recompression evaporator using steam jet recompressor.

a comparatively small work input is required; the compression step, when adiabatic or nearly so, raises the pressure and enthalpy simultaneously. The steam jet requires the use of high-pressure steam (around 100 pounds per square inch gage) to remove the vapor from the evaporator and force it into the steam chest at elevated pressure; with a steam jet, there is a surplus of vapor available.

Cremer and Davies<sup>10</sup> present a graph showing that 100 pounds per square inch gage steam is required to entrain 1 pound of vapor for various temperature "boosts" and various initial temperatures of the vapor. Their graph shows that the steam jet requirement increases with the magnitude of the temperature boost and with the initial temperature of the vapor.

Vapor recompression is not economical when evaporator temperatures and pressures are so low that the vapor volumes to be compressed are large. For one such case Cross<sup>12</sup> has shown that the latent heat from high-volume vapor may be transferred to a secondary vapor having a low specific volume (such as ammonia), which can then be compressed to provide the necessary temperature level for the evaporation process. The auxiliary equipment required for transferring the heat to and from the ammonia, however, makes this type of operation economically unattractive here.

#### COST OF ORDINARY RECOMPRESSION EVAPORATION

Consider the recompression evaporator shown in Figure 4. Suppose it receives as feed 1,000 pounds per hour of a solution containing 1 pound of contaminants per 999 pounds of water (1,000 ppm of contaminants) and discharges 950 pounds of distilled water per hour plus 50 pounds per hour of contaminant concentrate containing 1 pound of contaminants per 49 pounds of water. The concentrate will be evaporated to dryness in a separate operation. Let the boiling temperature in the evaporator be 215° F (15.59 pounds per square inch absolute pressure) and let the steam used as the heating medium condense at 223° F. The feed is assumed to be at 70° F and is preheated as shown in the figure in the feed heat exchanger.

From steam tables,<sup>18</sup> the enthalpy of the vapor from the boiling solution is 1,151.4 Btu per pound and its entropy is 1.7522. The pressure of the condensing steam at 223° F is 18.4 pounds per square inch absolute. If the vapor from the evaporator is adiabatically and isentropically compressed to 18.4 pounds per square inch, its temperature is 241.5° F and its enthalpy is 1,164.3. The steam is thus superheated by 241.5 - 223, or 18.5° F. The gain in enthalpy in the compressor (if 100% efficient) is 1,164.3 - 1,151.4, or 12.9 Btu per pound; if the compressor efficiency is 80 percent, however, its work input will be 12.9/0.8, or 16.1 Btu per pound. Some of this excess work will be dissipated as heat to the metal parts of the compressor, and some will be passed on to the compressed steam. Assume the total enthalpy gain of the steam in the 80 percent efficient compressor to be 14.0 Btu per pound. Then the enthalpy of the steam to the steam chest is 1,165.4, its temperature is 243° F, and its pressure is 18.4. The enthalpy lost by each pound of steam condensing in the steam chest at 223° F is 1,165.4 - 191.2, or 974.2; the latent heat of evaporation at 215° F is only 968.3 Btu per pound, so that the difference, 5.9 Btu per pound, may be charged to heat loss or to preheating.

Hence this unit, once it is brought to operating condition by auxiliary heaters, requires as its only energy input the work to the compressor of 16.1 Btu per pound of distilled water produced. If the efficiency of the electric motor driver is 90 percent, the energy requirement is 43.6 kilowatt-hours per 1,000 gallons of distilled water. With electricity at \$0.008 per kilowatt-hour, the electricity cost per 1,000 gallons of distilled water is \$0.35. This cost is too high to make the considered case competitive with those described on pages 17 to 19, and 25 and 26, but the power cost can be reduced if the temperature difference is made lower than the value of  $8^{\circ}\text{F}$  used in the example. Nevertheless, this design will not be optimized because of the comparatively low coefficients of heat transfer obtained in short-tube, natural-circulation evaporators. Although these units have been widely employed in smaller units, especially on shipboard for production of potable water from sea water (the so-called Kleinschmidt still), other equipment types are more economical for evaporation of large quantities of water in a recompression evaporator. These types are discussed later.

If a steam jet is used in place of the compressor for this process as in Figure 5, the steam requirement may be estimated by extrapolation from the data of Cremer and Davies.<sup>10</sup> With vapor initially at  $215^{\circ}\text{F}$ , and with a temperature boost of  $8^{\circ}\text{F}$  to  $223^{\circ}\text{F}$ , 0.47 pound of 100-pounds-per-square-inch-absolute steam is required to the steam jet per pound of vapor compressed. If 50 pounds of the 1,000 pounds of feed is to be left unevaporated as a contaminant concentrate, then there must be 950 pounds of vapor leaving the evaporator body. Let  $V$  be the pounds of excess vapor; then the vapor to be recompressed is  $(950 - V)$ , and the quantity of 100 pounds-per-square-inch-absolute steam required is  $0.47(950 - V)$ . The vapor quantity to the steam chest is  $1.47(950 - V)$ . It is now necessary to make an enthalpy balance around the evaporator to determine  $V$ :

#### Enthalpy

|                                    |                |           |
|------------------------------------|----------------|-----------|
| of input feed . . . . .            | (1,000)        | (182)     |
| lost by condensing vapor . . .     | 1.47 (950 - V) | (963.2)   |
| of vapor from evaporator . . . . . | (950)          | (1,151.4) |
| of liquid product . . . . .        | (50)           | (183)     |

Equating the first two items (input quantities) to the last two items (output quantities) reveals that  $V = 299$ , and the steam requirement to the jet is 306 pounds. With steam at 35 cents per 1,000 pounds, the steam cost is 93 cents per 1,000 gallons of distilled water, with no credit allowed for the 299 pounds of

excess vapor produced at 215° F and 15.6 pounds per square inch absolute pressure. Obviously considerable credit should be allowed for this steam, since it would be a useful heat source in a multiple-effect evaporation scheme.

Considerable effort is required to optimize a design using a steam jet recompressor and this will not be attempted. A combined mechanical recompression evaporator and multiple-effect evaporator, mentioned on page 15, produced water at a favorably low cost. It is likely that a similar combination process using a steam jet recompressor would have similar but not lower costs.

### COST OF A RECOMPRESSION-FLASH EVAPORATION CYCLE

A typical process utilizing vapor recompression combined with flash evaporation is shown in Figure 6. In this scheme,

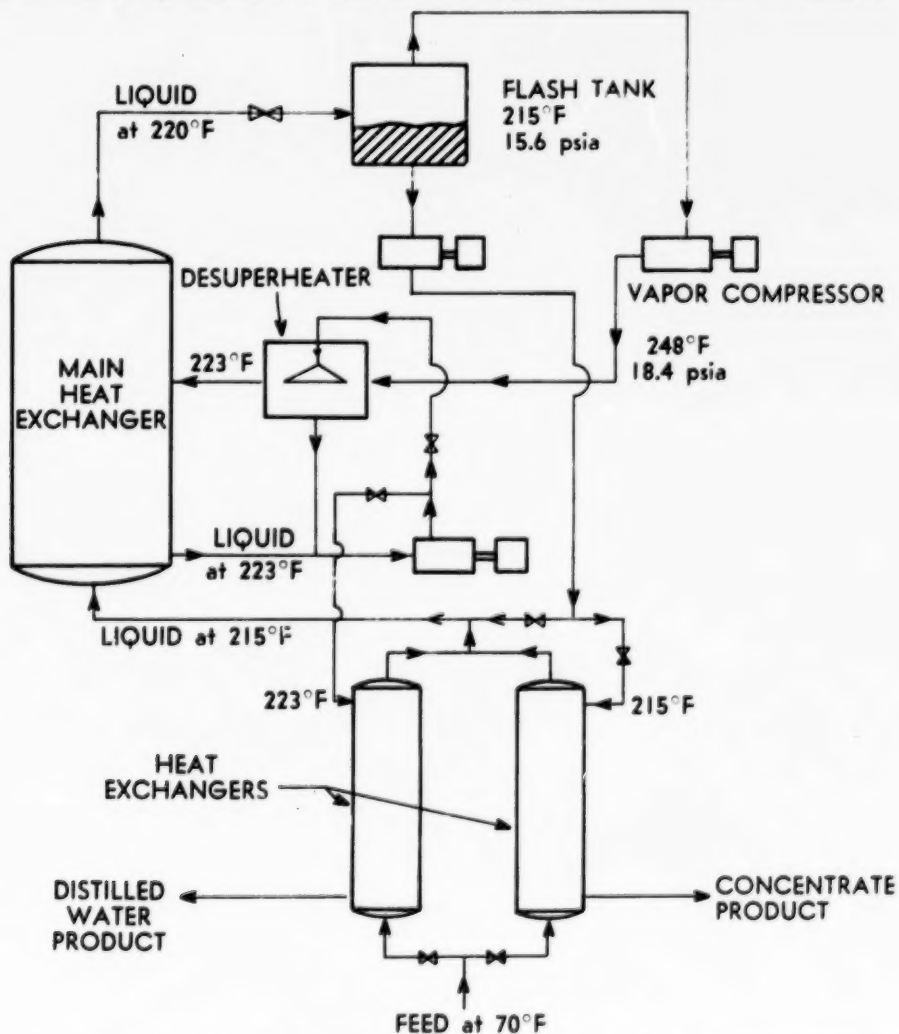


Figure 6. Production of distilled water by vapor recompression, flash evaporation process.

developed by Dodge and Eshaya,<sup>13</sup> the liquid to be evaporated does not boil inside the tubes, but is pumped through the tubes as a liquid at high velocity, and then lowered in pressure in a flash tank where partial evaporation occurs.

Typical operating conditions similar to those described on pages 29 to 31 are presented in the figure. The advantage of this type of operation lies with the high coefficients of heat transfer obtained.

The energy requirement in the recompression step of this process is computed as on page 29. The desuperheater shown on the flow sheet lowers the temperature of the vapor from the compressor from 248 to 223° F by contacting it with the condensate at 223° F; an enthalpy balance on the desuperheater shows that this produces 1.014 pounds of vapor for the main heat exchanger per pound of vapor from the compressor. The advantage of producing saturated vapor in the desuperheater is that all the comparatively expensive heat transfer area in the main exchanger can be used for condensation where coefficients are high, while the desuperheating can occur at low cost in the direct-contact unit.

Consider now the calculation of the total circulation rate through the flash tank and assume that the conditions shown on Figure 6 are followed. The enthalpy of 1 pound of liquid water at 220° F is 188.1 Btu per pound whereas the enthalpy of liquid water at 215° F is 183.1 and that for steam at 215° F is 1,151.4. An enthalpy balance on the flash tank shows that 0.00516 pound of vapor is formed there per pound of liquid entering; thus 194,000 pounds of liquid needs to be recirculated through the flash tank for every 1,000 pounds of fresh feed. This circulation rate is large, but the pressure differential across the valve before the flash tank is only 17.19 - 15.59, or 1.60 pounds per square inch, and the pressure drop through the exchanger could be kept to about the same value, so that the pumping requirement is not excessive. Dodge and Eshaya<sup>13</sup> evaluated the total energy requirement for this process for desalting sea water as follows.

Table 9. ENERGY REQUIREMENTS FOR DESALTING  
SEA WATER BY RECOMPRESSION FLASH  
EVAPORATION

|                                                    |                  |
|----------------------------------------------------|------------------|
| Brine velocity in main exchanger, ft/sec . . . . . | 8                |
| Fraction of brine flashed on one pass, % . . . . . | 0.005            |
| Mean $\Delta t$ in main exchanger, °F . . . . .    | 2.83             |
| Temp of brine leaving exchanger, °F . . . . .      | 212              |
| Compressor pressure ratio . . . . .                | 1.15             |
| Compressor efficiency, % . . . . .                 | 80               |
| Compression work, kw-hr/1,000 gal . . . . .        | 38.5             |
| Pumping work, kw-hr/1,000 gal . . . . .            | 8.0 <sup>a</sup> |
| Total work, kw-hr/1,000 gal . . . . .              | 46.5             |

<sup>a</sup> This figure depends upon piping arrangement, pump efficiency, and friction losses. It is a rough figure, but based upon many detailed calculations and is believed to be a reasonable one.

The conditions chosen in Table 9 are similar to those in Figure 6 for the distillation of waste water. The total work of 46.5 kilowatt-hours per 1,000 gallons corresponds to an expenditure of 37 cents per 1,000 gallons with electricity at \$0.008 per kilowatt-hour.

The work requirement for any vapor recompression process is a linear function of the temperature difference across the evaporating surface, at least up to a  $\Delta t$  of 20° F. Thus if values of  $\Delta t$  less than the 8° F considered above are used, energy costs could be lowered. Optimization of the TTD for the multi-stage flash evaporation of sea water gave 4° F (page 24); if this value is used, then electricity costs would be about one-half of 37 cents per 1,000 gallons or 19 cents per 1,000 gallons. If costs other than power costs (mainly equipment costs) can be kept to a similar figure — say 19 cents per 1,000 gallons, then the total water cost by the scheme of Figure 6 would be 38 cents per 1,000 gallons, a figure slightly greater than those obtained when distillation is carried out by multi-stage flash or by LTV, multiple-effect evaporation.

Actually, the recompression-flash evaporation scheme is quite similar to the multi-stage flash scheme, since in both cases the main heat transfer operation is between condensing steam on the shell side and a high-velocity liquid stream on the tube side. Temperature differences, heat transfer coefficients, and heat loads are similar in the two cases, so the tube areas should be similar. The main difference is that in the recompression-flash evaporation scheme a large amount of liquid is recirculated through a short length of tubes with low pressure

drop, while in the multi-stage flash scheme a smaller amount of liquid is recirculated through a longer length of tubes with higher pressure drop. These effects counterbalance each other so that pumping costs in the two cases should not be greatly different, nor should the equipment costs. Costs other than steam for multi-stage flash evaporation were shown to be

$$\frac{51.34 - 22.49}{1.90} = 15.2 \text{ cents per 1,000 gallons (see Table 8), and}$$

this figure is then a good approximation of similar costs for recompression-flash evaporation. With power costs for recompression-flash evaporation computed above as 19 cents per 1,000 gallons, the total water cost by this process may be estimated as  $15.2 + 19$  or 34 cents per 1,000 gallons.

#### COMPARISON OF RECOMPRESSION EVAPORATOR WITH MULTIPLE-EFFECT EVAPORATOR

Dodge and Eshaya<sup>13</sup> have compared the relative energy cost for a recompression evaporator with that for a multiple-effect evaporator. The work of compression,  $W$ , for a recompression evaporator, expressed as kilowatt-hours per 1,000 gallons of product is:

$$W = \frac{0.1345 T_1 (r^2 - 1)}{\epsilon r} \quad (1)$$

where  $T_1$  is the compressor intake temperature, degree R ;  $r$  is the pressure ratio in the compressor; and  $\epsilon$  is the adiabatic efficiency of the compressor. For the multiple-effect evaporator, if  $w_1$  is the pounds of water evaporated per pound of steam in each effect, the total pounds evaporated per pound of steam in  $n$  effects, designated as  $\Sigma w$ , is

$$\Sigma w = \frac{w_1 (1 - w_1^n)}{1 - w_1} \quad (2)$$

The cost of energy for each case may now be equated, with the result that  $n$ , the equivalent number of effects required to give the same energy cost as the recompression evaporator, is

$$n = \frac{\log \left[ 1 - \frac{61.8(1-w_1)r\epsilon C_s}{w_1 T_1 (r^2 - 1) C_E} \right]}{\log w_1} \quad (3)$$

where  $C_s$  is the cost of steam, cents per 1,000 pounds, and  $C_E$

is the cost of electricity, cents per kilowatt-hour.

In applying Eq (3) to the recompression evaporation system shown in Figure 4, where the average  $\Delta t$  in the main exchanger is  $5.1^{\circ}\text{ F}$ , and where the liquid temperature into and out of the flash chamber differs by  $5.0^{\circ}\text{ F}$ , the value of  $r$  is  $18.4/15.6$  or  $1.18$ ;  $\epsilon$  is taken as  $0.80$ ,  $T_1$  as  $675^{\circ}\text{ R}$ , and  $C_E$  is taken as  $0.80$  cents. If the multiple-effect evaporator operates with a  $\Delta t$  of  $5^{\circ}\text{ F}$  in each effect, and if the boiling-point rise of the solution being evaporated is negligible, then it may be assumed that the pounds of water evaporated per pound of steam in each effect,  $w_1$ , is equal to the ratio of the latent heats of vaporization on each side of the heat transfer surface (neglect heat losses and sensible heats of the liquid). As latent heats increase with decreasing temperature, this ratio is less than unity, but with only a  $5^{\circ}\text{ F}$  difference in boiling temperature, it is close to unity:  $0.9973$  at  $70^{\circ}\text{ F}$  and  $0.9968$  at  $215^{\circ}\text{ F}$ . If  $w_1$  is taken as  $0.9970$ , the average of these latter two figures, then substitution into Eq (3) shows that  $n = 15.8$ . Thus, even under the favorable conditions of steam economy chosen for the multiple-effect evaporator, the recompression evaporator will produce  $1,000$  gallons of distillate for the same energy cost as a multiple-effect evaporator if the latter contains  $15.8$  effects. This calculation indicates that recompression evaporation is competitive in terms of cost with other processes employing re-use of steam.

It is of interest to compute  $n$  for other conditions of  $w_1$ , since this value will vary with  $\Delta t$  and with heat losses (Table 10).

Table 10. COMPUTATION OF  $n$  FOR OTHER CONDITIONS OF  $w_1$

| $w_1$            | $n$  |
|------------------|------|
| 0.9500 . . . . . | 27.8 |
| 0.9900 . . . . . | 16.5 |
| 0.9970 . . . . . | 15.8 |
| 0.9990 . . . . . | 15.5 |

This table shows the insensitive nature of the result in the range  $0.990$  to  $0.999$ .

HICKMAN RECOMPRESSION STILL

Considerable effort has been expended by the Office of Saline Water on development of the Hickman recompression still.<sup>16</sup> In this apparatus a number of rotors are fed with sea water on one surface: the feed flows outward in a thin film at high velocity

because of the centrifugal action; steam condensing on the other side of the 0.064-inch-thick copper evaporation surface vaporizes a portion of the feed. The heat transfer coefficients range from 1,600 to 2,600 Btu per hour per square foot per  $^{\circ}$  F. The steam evolved from the evaporation of the feed liquid is led to a motor-driven centrifugal compressor, after which it is used as the heating medium. The steam condenses on the opposite side of the rotors, and the condensate is flung outwards and collected separately from the unevaporated portion of the feed.

Tests with a still of this type designed to produce 25,000 gallons of distillate per day were completed in 1958.<sup>16</sup> Some tests were carried out with fresh-water feed, which is of interest in the present case. The best performance with fresh water produced 25,200 gallons per day of distillate from 68,520 gallons per day of feed with a condensing temperature of  $135.2^{\circ}$  F and a temperature difference of  $5.23^{\circ}$  F. The heat transfer coefficient was 2,610, the total kilowatt-hours per 1,000 gallons for the installation was 51.4, and the kilowatt-hours per 1,000 gallons for the compressor only was 34.3. Under some conditions the rotors were unevenly wet or incompletely wet; noncondensable gases were also found to be present under some conditions. Tests showed that as little as 0.1 percent of foreign gas in the steam reduced the rate of distillation by 30 percent. Both of these difficulties are, however, capable of being remedied, according to the authors.<sup>16</sup>

Subsequent tests made with a flat-plate rotor 4-1/2 feet in diameter and fed with eight feed-water nozzles showed that heat transfer coefficients of 2,800 could be obtained with a rotor speed of 300 rpm. The coefficients increased almost linearly with rotor speed, reaching 4,500 at a speed of 750 rpm.

Hickman et al.<sup>16</sup> estimated the cost of producing potable water from sea water by a 30-rotor still designed to produce 100,000 gallons per day. The feed-distillate ratio was 2.5 to 1, and the rotor speed was 400 rpm. The heat transfer coefficient was taken as 3,000 at  $110^{\circ}$  F, and it was increased 5 percent with every  $10^{\circ}$  F increase in evaporating temperature. The rotating assembly was assumed to have a useful life of 5 years, and all other components were estimated to have a 20-year life. At any particular evaporating temperature, a plot of cost versus temperature difference reached a minimum value between 6 and  $8^{\circ}$  F; minimum operating costs were found to be \$1.28, \$1.31, and \$1.40 per 1,000 gallons for evaporating temperatures of 150, 120, and  $90^{\circ}$  F, respectively. The capital costs were \$1.70, \$1.72, and \$1.80 per gallon per day.

Since these costs are somewhat higher than costs for other distillation processes, they will not be recomputed for a waste-water feed, since a change in type of feed could not alter the cost sufficiently to make this process competitive.

#### MULTIPLE-EFFECT, ROTATING EVAPORATOR

The Hickman still described on pages 35 to 37 operates with a single temperature (and pressure) level for the evaporating liquid and uses vapor recompression to obtain its steam economy. A different type of rotating evaporator, studied by Clark and Bromley,<sup>9</sup> contains a number of closely spaced, rotating plates. The pressure between each pair of plates is made successively lower so that the vapor produced from a lower plate becomes the heating medium for the liquid flowing outward on the plate above; the pure condensate is held on the underside of the rotating plate by surface tension and is driven to the outer edge by centrifugal force. The unit obtains its steam economy by the same principle as a multiple-effect evaporator. Provisions are included for preheating the feed in an exchanger at the periphery of the evaporator from which the feed streams to each plate are drawn at temperatures approximating the evaporating plate temperatures. About 50 percent of the feed is evaporated while it flows across the plate in a thin film.

These authors present operating data for a 4-foot-diameter unit containing 29 effects (or rotating plates). The unit produced about 3,000 pounds of distillate product per hour when the rotor speed was 970 rpm. The plates were 9/16 inch apart. The steam economy was 23.1 pounds of product per pound of steam and the measured over-all coefficient of heat transfer was around 3,000 Btu per hour per square foot per ° F. The preheater approach temperature was about 7° F. Their calculations show that the steam economy increases almost linearly with the number of effects; when 50 effects are used, 37 pounds of product can be made per pound of steam. Increasing the number of effects has very little effect upon the product rate. When the rotating speed or evaporator diameter is increased, the product rate is increased without any appreciable change in economy. Lowering the preheater approach temperature increases the steam economy, while a decrease in the boiling-point elevation — an important factor in applying this type of unit to purification of water containing less than 0.1 percent of solids — improves considerably both the economy and the product rate.

It is difficult to make cost estimates for this type of apparatus since it has a highly specialized design. No cost data are given in the reference cited. The economic possibilities do not

appear so overwhelmingly great, however, that costs from this apparatus would be appreciably lower than costs obtained from use of the more standard methods of distillation.

## CONCLUSION

Of the many distillation and evaporation processes shown in this report to be feasible for purification of municipal waste waters, three show cost advantages over the others. These are multi-stage flash evaporation, multiple-effect evaporation, and recompression-flash evaporation. A combination of the latter two processes has a similar cost advantage.

In a consideration of any of these processes, it should be remembered that pretreatment of the raw feed will probably be needed to remove at least some of the dissolved bicarbonates to prevent excessive fouling. If preremoval of about 100 ppm of inorganic contaminants by mixed-bed ion exchange is sufficient to reduce fouling to an acceptable level, this pretreatment would raise the cost of purified water by about 7 cents per 1,000 gallons (see page 9). This figure is roughly 15 percent of the cost of carrying out the various distillation processes.

The vapor re-use processes described above recover 95 percent of the original feed as a distilled water product, but consideration must also be given to disposal of the remaining 5 percent of the original feed, which is a concentrate containing 2 percent contaminants. In evaporation of this concentrate nearly to dryness, the solids increase to a very large concentration, which greatly affects corrosion, fouling, and boiling-point rise. Also, owing to the unknown composition and variable concentration of the contaminants, a lower degree of automatic control and greater flexibility in operations will be required; submerged-combustion techniques may be useful in this application. These factors will increase both equipment and steam costs; even in a large scale of operation, the total cost may be considered to be three times greater per gallon of feed than in the main evaporation step. With only 50 gallons to be treated per 1,000 gallons of original feed, however, this final evaporation step raises the water purification cost by a factor of only  $(50/1,000) (3) = 0.15$  or 15 percent.

It is now possible to estimate the over-all cost of purifying contaminated waters by distillation. The three comparatively low-cost vapor re-use processes described above are seen on pages 18, 26, and 33 to produce distilled water at an expense

that is from 9 to 35 percent less than the expense of producing distilled water from sea-water feed; if the most favorable of these is taken, and the costs of ion exchange pretreatment (15 percent) and disposal of the 2 percent contaminant concentrate (15 percent) are added, the over-all cost of purifying municipal waste waters is about the same as that for separating a sea-water feed into equal parts of distilled water and a 7 percent concentrate.

It is not possible at this time to set down an absolute cost figure for purification of either a 1,000 ppm waste water or a 3.5 percent salt concentration of sea water. The cost analyses contained in the body of this report are based upon cost information from several years ago and for plants having capacities that may or may not be consistent with the size of plants projected for waste-water purification. Nevertheless, cost experience gained by the Office of Saline Water's demonstration plants has an important bearing on the present problem. Chemical and Engineering News of March 12, 1962 (p. 24), reports that the San Diego, Calif., multi-stage flash demonstration plant, built at a cost of \$1.6 million to produce 1 mgd of potable water, is converting water for about \$1.14 per 1,000 gallons of product. Improvements are planned to reduce this cost by at least 10 cents per 1,000 gallons, and a speculative figure of 75 cents per 1,000 gallons is given as the cost of producing water by the same process in a 10- to 15-million-gallon-per-day plant. The same figure might then be applied speculatively as a recent value for the cost of purifying waste water in a similarly sized plant.

It is well to recall the assumptions made in this report and in the cost calculations of the Office of Saline Water. These are: (1) That with suitable pretreatment, waste water can be evaporated at temperatures around 350° F without excessive fouling, (2) that steel equipment can be used, and (3) that maintenance costs and useful life of the plant (20 years) have been correctly estimated.

Another important factor should be re-emphasized; it deals with the use of exhaust steam from turbines employed in electric-power generation as the energy source for multi-stage flash and multiple-effect evaporators. It has been assumed that much of the cost of the original steam can be charged against the power generation. This assumption is valid only if loading of the evaporators can be adjusted to coincide with the loading of the generators; whether this can actually be accomplished depends upon ability to store feedwater and other factors. If variable loading of the evaporators is a problem, use of thermocompression evaporators may be preferred since this equipment uses electric power (to drive the compressors) as the only energy input.

## REFERENCES

1. Bloemer, J. W., Collins, R. A., and Eibling, J. A., Field Evaluation of Solar Sea Water Stills, Saline-Water Conversion, Advances in Chemistry Series No. 27, p. 166, American Chemical Society, Washington, D. C., 1960.
2. Brice, D. B., Dusabek, M. R., and Townsend, C. R., Economics of Sea-Water Distillation in Southern California, Metropolitan Water District of S. Cal., Fluor CRR 1,056, Dec. 1959.
3. Brice, D. B., et al. Office of Saline Water R. and D. Progr. Report 34, Fluor CRR 1,046, OTS No. PB 161,010, Aug. 1959.
4. Brice, D. B., and Townsend, C. R., Sea-Water Conversion by the Multi-Stage Flash Evaporation Method, Saline-Water Conversion, Advances in Chemistry Series No. 27, p. 147, American Chemical Society, Washington, D. C., 1960.
5. Bunch, R. L., Barth, E. F., and Ettinger, M. B., Organic Content of Secondary Sewage Effluents, Robert A. Taft Sanitary Engr. Center, U. S. Department of Health, Education, and Welfare, Cincinnati 26, Ohio.
6. Chemical Eng. Progr., 54, No. 2, 87, 1958.
7. Chemical Eng. Progr., 56, No. 3, 81, 1960.
8. Chemical Eng. Progr., 57, No. 3, 94, 1961.
9. Clark, R. L., and Bromley, L. A., Chemical Eng. Progr., 57, No. 1, 64, 1961.
10. Cremer and Davies, Chemical Engineering Practice, 6, p. 113, Butterworths, London, 1958.
11. Cronan, R., Chem. Engineering, 63, No. 2, 163, 1956.
12. Cross, J. A., Chem. Eng. Progr., 54, No. 4, 132, 1958.
13. Dodge, B. F., and Eshaya, A. M., Thermodynamics of Some Desalting Processes, Saline-Water Conversion, Advances in Chemistry Series NO. 27, p. 7, Amer. Chem. Soc., Washington, D. C., 1960.
14. Eibling, J. A., and Hyatt, D. L., Chem. Eng. Progr., 54, No. 10, 84, 1958.

15. Florine, R. E., and Dedert, W. G., Chem. Eng. Progr., 54, No. 4, 64, 1958.
16. Hickman, K. C. D., et al. Centrifugal Phase-Barrier Recompression Distillation, Saline-Water Conversion, Advances in Chemistry Series No. 27, p. 128, Amer. Chem. Soc., Washington, D. C., 1960.
17. Hubbell, J. P., Chem. Eng. Progr., 54, No. 7, 20, 1958.
18. Keenan and Keyes, Steam Tables, John Wiley, New York, 1950.
19. Kobe, K. A., and Hall, R. N., Jr., Heating of Water by Submerged Combustion, Circular No. 24, Univ. of Texas; Bureau of Engr. Res., Austin 12, Tex., June 1954.
20. Kohlins, W. D., and Englander, H. P., Chem. Eng. Progr., 52, No. 2, 45-F, 1956.
21. Lof, G. O. G., Design and Operating Principles in Solar Distillation Basins, Saline-Water Conversion, Advances in Chemistry Series No. 27, p. 156, Amer. Chem. Soc., Washington, D. C., 1960.
22. Lof, G. O. G., Southwest Water Works Jour., 42, No. 4, 14, 1960.
23. Nelson, W. L., Oil and Gas Journal, 56, Mar. 17, 1958, p. 190.
24. Ibid., June 23, 1958, p. 137.
25. Ibid., June 30, 1958, p. 103.
26. Ibid., July 14, 1958, p. 131.
27. Othmer, D. F., et al. Chem. Eng. Progr., 57, No. 1, 47, 1961.
28. Saline Water Conversion Report for 1959, Office of Saline Water, U. S. Dept. of Interior, Washington, D. C., Jan. 1960.
29. Standardized Procedure for Estimating Costs of Saline-Water Conversion, Office of Saline Water, U. S. Dept. of Interior, Washington, D. C., Mar. 1956.

30. Standiford, F. C., and Bjork, H. F., Evaporation of Sea Water in Long-Tube Vertical Evaporators, Saline-Water Conversion, Advances in Chemistry Series No. 27, p. 115, Amer. Chem. Soc., Washington, D. C., 1960.
31. Strobel, J. J., Chem. Eng. Progr., 57, No. 1, 37, 1961.
32. Weisman, W. I., Application of Submerged Combustion in Industrial Waste Treatment, Proc. 8th Ind. Wastes Conf., Purdue Univ., 363, 1953.
33. Ind. Eng. Chem., 49, No. 12, 53A, 1957.
34. Clark, R. L., et al. Heat of Concentration and Boiling Point Elevation of Sea Water, Saline-Water Conversion, Advances in Chemistry Series No. 27, p. 21, Amer. Chem. Soc., Washington, D. C., 1960.
35. Fink, F. W., Corrosion of Metals in Sea Water, ibid., p. 27.
36. Nachod, F. C., and Schubert, J., Ion Exchange Technology, p. 203-218, Academic Press, New York, 1956.